



Continuous hydrino thermal power system

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ABSTRACT

The specifics of a continuous hydrino reaction system design are presented. Heat from the hydrino reactions within individual cells provide both reactor power and the heat for regeneration of the reactants. These processes occur continuously and the power from each cell is constant. The conversion of thermal power to electrical power requires the use of a heat engine exploiting a cycle such as a Rankine, Brayton, Stirling, or steam-engine cycle. Due to the temperatures, economy goal, and efficiency, the Rankine cycle is the most practical and can produce electricity at 30–40% efficiency with a component capital cost of about \$300 per kW electric. Conservatively, assuming a conversion efficiency of 25% the total cost with the addition of the boiler and chemical components is estimated at \$1064 per kW electric.

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1. System design

The multi-cell reactor to continuously generate power shown in Fig. 1 comprises a plurality of repeating planar layers of insulation, heat exchanger, and reaction cell. Each cell produces power continuously to elevate its reactant temperature higher than that required for regeneration. An exemplary reaction to form hydrinos is a hydride exchange between an alkali hydride catalyst and source of hydrogen, and an alkaline earth metal or lithium metal. The reactants, exchange reactions, products, and regeneration reactions and parameters were reported previously [1–7]. The initial alkali hydride is regenerated by evaporation at 400–550 °C and condensation at a temperature of about 100 °C lower in the presence of hydrogen that reacts to form the alkali hydride. Thus, a heat gradient exists between the reactants at an elevated temperature and a cooler zone in each cell that drives the thermal regeneration. The cells are horizontally oriented with a dead space along the longitudinal axis of the cell that allows the alkali metal vapor to escape from the reactants along the bottom of the cell during continuous regeneration. The metal condenses in the cooler zone along the top of the cell. The cooler region is maintained at the desired condensation temperature by a heat collector comprising heat transfer tubes with a variable heat acceptance rate at the top of each cell. The heat exchanger comprises a water wall of heat transfer tubes with flowing water heated to steam. Specifically,

saturated water flows through the heat transfer tubes, absorbs energy from the cells, and evaporates to form steam.

Thus, the heat is ultimately transferred to water that is boiled in tubes peripherally to each reactor cell wherein the heat transfer tubes form a water wall. The temperature of the boiling water must be in the much lower temperature range of 250–360 °C. These temperatures are high enough to achieve nucleate boiling, the most effective means of heat transfer to a water medium; but are below the ceiling set by the excessive steam pressures at temperatures above this range. The nucleate boiling of water occurs on the inner surface of each heat transfer tube wherein an even temperature distribution in the water wall is maintained due to the tubes being embedded in the highly conductive thermal medium such as copper, and additionally the water that was not evaporated to steam is recirculated. Heat flows from the top cell wall through the medium to the heat transfer tubes. Due to the required much higher temperatures in each cell even at the lower end of its gradient, a second temperature gradient is maintained between each cell top and the heat load, the boiling water and subsequent systems. Since the heat transfer tubes have a higher capacity to remove heat than each cell has to produce it, a second external thermal gradient is maintained by adding one or more thermal barriers between the top-half of the cell wall and the water wall. The desired high internal cell temperatures as well as the gradient are achieved by insulating at least one of the top-half of the cell and the outer wall of each heat transfer tube from the conductive medium. The cell temperatures and gradient are controlled to optimal values through the variable heat transfer by adjusting the thermal barriers at the top-half of the cell and the

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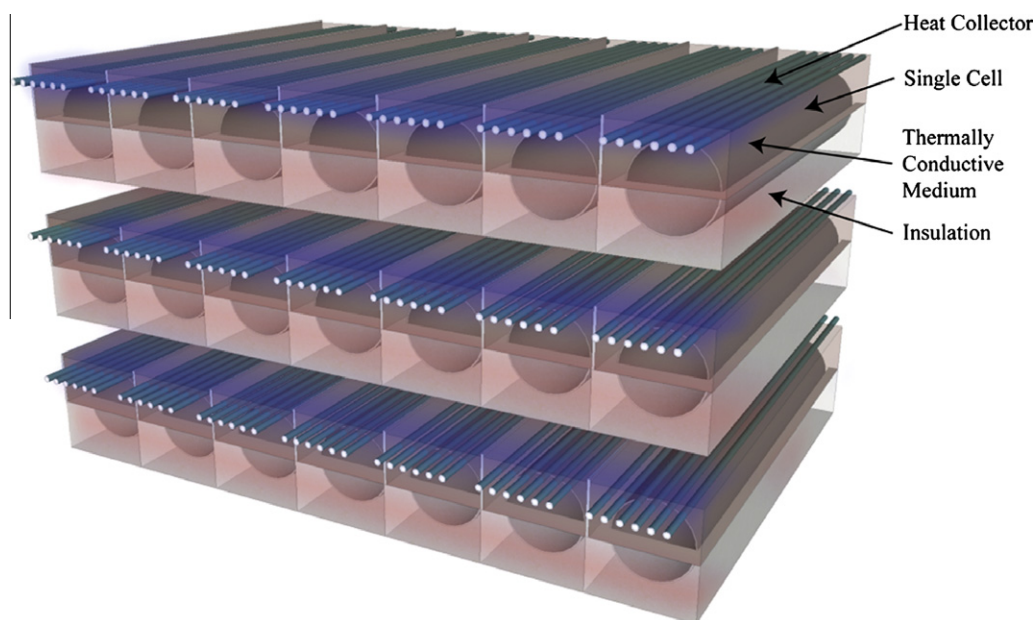


Fig. 1. A schematic drawing of a multi-cell reaction system comprising alternate layers of insulation, reactor cells, and heat exchanger to maintain continuous power via a cell heat gradient. The reactant alkali hydride is continuously regenerated by product decomposition and alkali metal evaporation in the elevated-temperature bottom zone maintained by the reaction with condensation and rehydriding in a cooler top zone maintained by the heat collector. A rotating wiper blade rejoins the regenerated alkali hydride with the reaction mixture.

heat transfer tubes, the thermal conductivity of the medium penetrated by the heat transfer tubes, the heat exchanger capacity and the steam flow rate in the tubes. In the former case, the thermal barriers may each comprise a gas or vacuum gap that is variable based on the gas composition and pressure. The multi-cell reaction system is assembled into a boiler system shown in Fig. 2 to output steam. The steam generated in the heat transfer tubes of the water wall may flow to a turbine and generator to produce electricity directly, or the water wall may feed steam into a primary steam loop that transfers heat to a secondary steam loop through a heat exchanger. The secondary loop may power a turbine and generator to produce electricity. A steam turbine receives the steam from the boiling water, and electricity is generated with a generator as shown in Fig. 3.

After the condensed metal such as K or Na is hydrided due to the presence of hydrogen in the cell including replacement hydrogen for that consumed to make hydrides, the hydride is returned to the bottom of the cell and mixed with the other reactants. One or more internal rotating wiper blades or stirrers may be swept along the inner cell wall to mix the formed hydride with the other reactants. Optionally, rejoining of the alkali hydride with the other reactants and chemical mixing is achieved by rotating the cell about its longitudinal axis. This rotation also transfers heat from the bottom position of the cell to the new top position following rotation; consequently, it provides another means to control the internal cell temperature gradient for alkali metal transport. However, as shown in Appendix B, the corresponding heat transfer rate is high requiring a very low rotational rate to maintain the heat gradient. The mixing rotation of the wiper blades or cells may be driven by an electric motor wherein the cells may be synchronized using gearing. The mixing may also be by magnetic induction through the cell wall of low permeability such as one of stainless steel.

2. System operating parameters

To achieve a cell-bottom temperature in the range of 400–550 °C on the higher-temperature power generation side of the

gradient and a temperature of about 100 °C lower at the regeneration side at the top, the cells have a heat collector only at the top as shown in Fig. 1, the power-producing reactants are located in the bottom, and the bottom section of the cell is insulated. The selected system design parameters are the (1) cell dimensions, (2) number of cells in the system, (3) the thermal resistance of the material surrounding the bottom half of the cell, (4) the thermal barrier at the top-half of the exterior wall of the cell, (5) the thermal conductivity of the medium surrounding the top-half of the cell that is penetrated by the heat transfer tubes, (6) the thermal barrier at the exterior heat transfer tubes' wall, (7) the heat transfer tubes' number, dimensions, and spacing, (8) the steam pressure, and (9) the steam flow and recirculation rates. The system design parameters are selected to achieve or maintain the desired operating parameters of (1) temperature and internal and external temperature gradients of each cell, (2) temperature of the boiling water at the periphery of the power flow from the cells, and (3) adequate boiling surface heat flux. Reaction parameters for the design analysis can be obtained experimentally on the various possible hydride exchange reactions that result in the formation of hydrides with significant kinetics and energy gain as well as comprising reactions that can be thermally regenerated. The power and regeneration chemistries and their parameters were reported previously [1–7]. Typical operating parameters for design engineering purposes are 0.25 W/cc constant power, 0.67 W/g reactants, 0.38 g/cc reactant density, 50 MJ/mole H_2 , 2–1 energy gain relative to hydride regeneration chemistry, equal reaction and regeneration times to maintain constant power output, and temperatures of 550 °C and 400–450 °C for power and regeneration, respectively, wherein the reaction temperature is sufficient to vaporize the alkali metal at the cell bottom, and the internal thermal gradient maintains the regeneration temperature at the cell top. Using the reactants and power densities, the reactant volume and total mass of the reactants to generate 1 MW of continuous thermal power are 3940 l and 1500 kg, respectively. Using a 0.25% reactant fill factor, the total reactor volume is 15.8 m³.

In the sample design summarized from the calculations given in Appendix A, the boiler comprises 140 stainless steel reaction

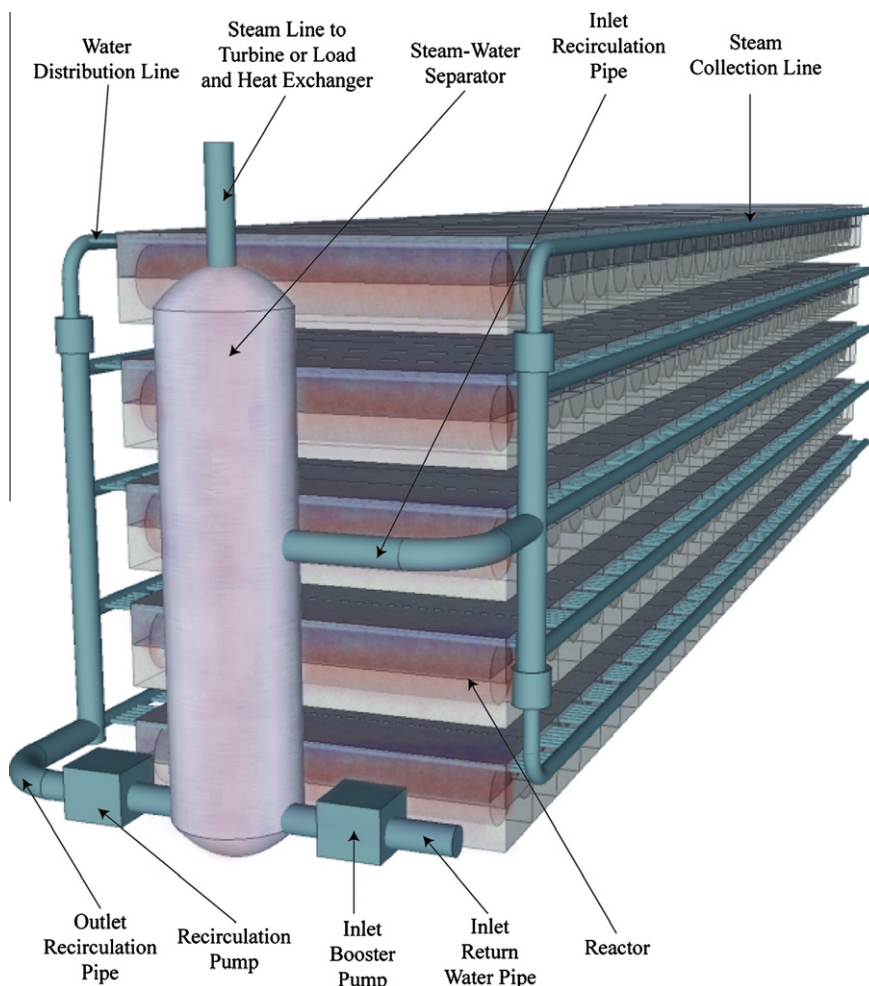


Fig. 2. The boiler system comprises the multi-cell reaction system shown in Fig. 1 and a coolant (saturated water) flow regulating system. The multi-cell reaction system heats the saturated water and generates steam. The flow regulating system collects the flow of saturated water, separates the steam and water, recirculates the separated water through the heat transfer tubes, and outputs and channels the steam into a main steam line. Insulation on all pipes not shown.

cells having a 176 cm length, 30.5 cm OD, a 0.635 cm cylindrical wall thickness, and 3.81 cm thick end plates. The wall thickness meets the design requirements for an internal pressure of 2.27 MPa at 550 °C due to the equilibrium decomposition pressure of the pressure-determining reactant NaH. Each cell weighs 120 kg and outputs 7.14 kW of thermal power. The bottom half of each cell is embedded in insulation. Copper or aluminum shot, a highly thermally conductive medium that is penetrated with the heat transfer tubes, surrounds the top-half of each cell. The temperature within the cell ranges between about 550 °C at the bottom wall to 400 °C at the wall surface facing the shot. As shown in Fig. 5, the 30.5 cm OD cross-sectional span of each cell is covered by six, 2.54 cm OD heat transfer tubes with a thickness of 0.32 cm that are evenly spaced at 5.08 cm centers. The heat flux at the internal surface of each heat transfer tube is about 11.8 kW/m² that maintains the temperature of each heat transfer tube external surface at about 367 °C. As shown in the flow diagram of steam generation (Fig. 4), water at room temperature (about 25 °C) flows into a heat exchanger where it is mixed with saturated steam and heated to a saturated temperature of 360 °C by the condensation of steam. A booster pump increases the water pressure to a saturation pressure of 18.66 MPa at 360 °C. The saturated water flows into the steam generator (comprising the water wall) to generate steam at the same temperature and pressure. Part of the steam flows back to the heat exchanger to preheat incoming return water from a turbine, while part of it

goes to the turbine to generate electrical power. Additionally, the non-evaporated water in the water wall is recirculated to maintain an even temperature along each heat transfer tube. To achieve this, a steam collection line receives steam and non-evaporated water and delivers it to a steam-water separator (Fig. 2). Water is pumped from the bottom section of the separator to return to the heat transfer tubes through a water distribution line. The steam flows from the top of the separator to the turbine with a fraction diverted to the heat exchanger to preheat the return water from the turbine. The saturated water flow rate from the 140-cell reactor system is 2.78 kg/s in the heat transfer tubes, and the total steam output flow rate is 1.39 kg/s.

3. Costs

To a good approximation, the total cost of the power system is the sum of the cost of the major components such as the (i) chemicals, (ii) the boiler comprising the reactor cells, the hydrogen pump, heat transfer tubes, steam generator, and the boiler housing, and (iii) the balance of the plant comprising the thermal to electric conversion system (including the feed water system, steam turbine generator, instrumentation and controls, electrical accessory switchgear, and other miscellaneous balance of plant systems). The total volume to produce 1 MW thermal is 15,800 l, corresponding to 0.25 W/cc constant power in the reactant volume of 3947 l. The reaction power per weight is 0.67 kW thermal (kWt)/kg

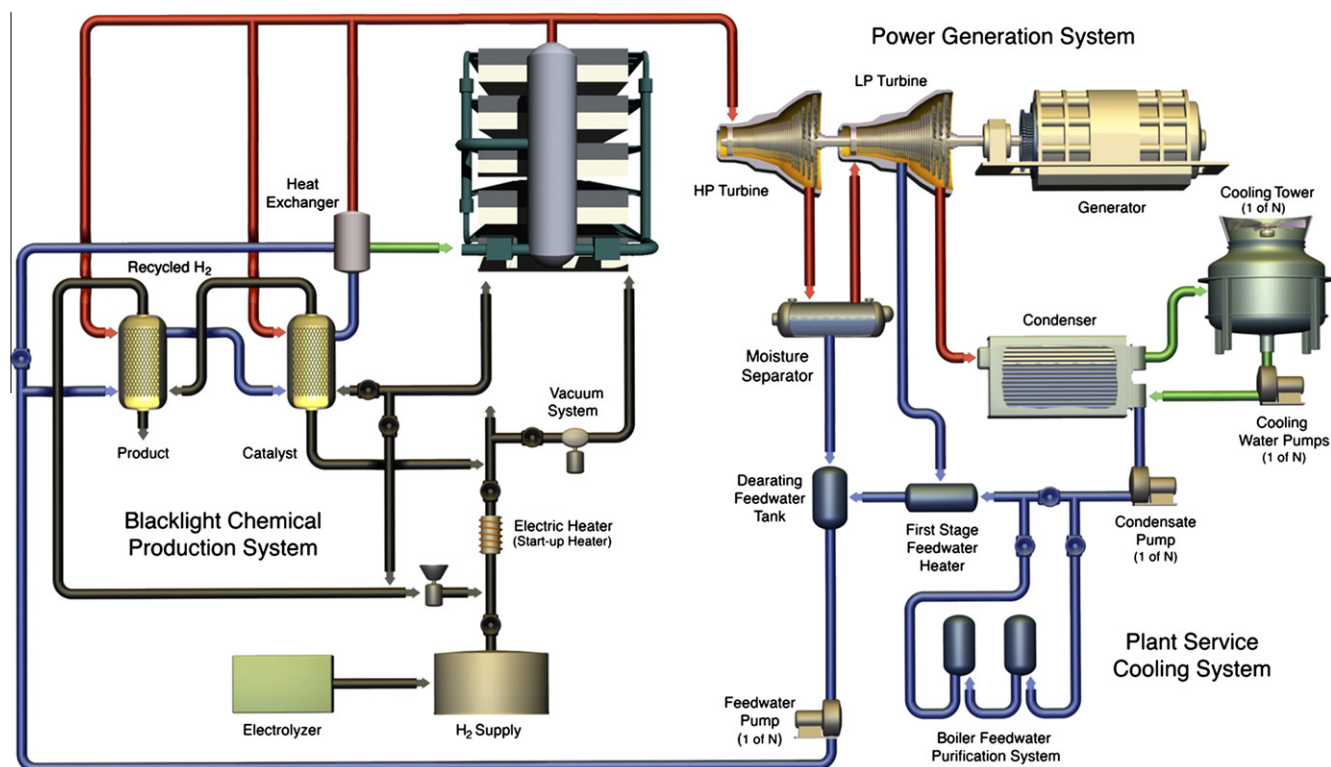


Fig. 3. Plant process schematic. Steam is generated in the boiler system and output from the steam-water separator to the main steam line. A steam turbine receives the steam from boiling water, and electricity is generated with a generator. The steam is condensed and pumped back to the boiler system.

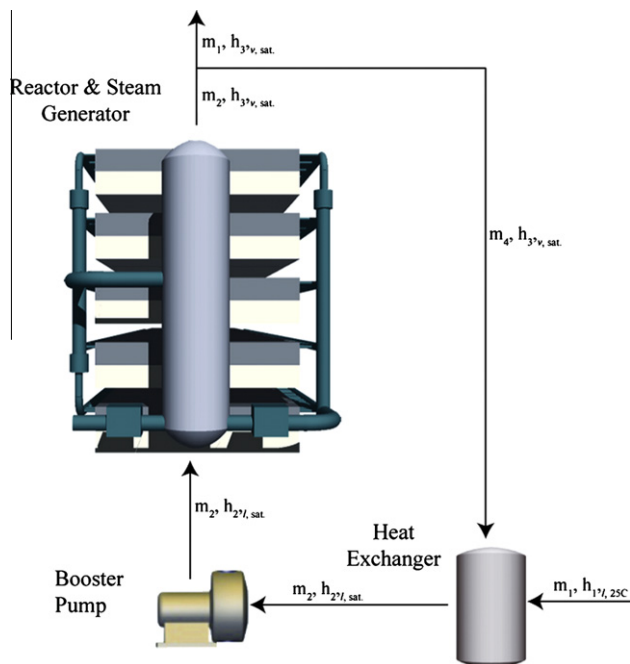


Fig. 4. Steam generation flow diagram.

reactants. The reactants of a typical reaction mixture of NaH, Mg, and TiC are in the weight percentages of 28, 28, and 44, respectively. Using the costs per commodity (Table 1) and the wt.%, the cost of the reactant mixture is \$17 per kWt and \$68 per kW electric (kWe) (Table 3).

From Table 2, the cost of the boiler for the 1000 kWt/250 kWe system is \$174,529 corresponding to \$174 per kWt and \$696 per

Table 1

Cost of commodities constituting the compounds of the reaction mixture [1–3].

Element	Cost (US \$/kg)
K	2 ^a
Na	1.6 ^a –2.5 ^{b,c}
Li	6.5 ^a
Mg	2.5 ^d
Carbon	1.5–11 ^e
Ti	2.2 ^d (23.5 for TiC ^f)
Y	35 ^g
Ca	2 ^b
Sr and Ba	6 ^h
Cu	6.6 ⁱ
Cl ₂	0.5 ^b
Br ₂	2 ^{b,j}

^a Source <http://metalsplace.com/news/articles/27731/china-market-prices/> (accessed on January 25, 2010).

^b Average price per kg from US import and/or export data based on annual reports from 2007, 2008 and thru November 2009, <http://dataweb.usitc.gov> (accessed January 26, 2010).

^c Email quote from DuPont, January 29, 2010.

^d Platts Metals Weekly (accessed December 21, 2009).

^e Calgon Carbon Corporation Quote, \$1.5 for bulk orders on January 25, 2009.

^f Inframat Advanced Materials Quote, February 2, 2010.

^g Metal Prices Report dated 26 November 2009 provided by Metal Pages.com as an email service.

^h Average price per kg from US import and/or export data based on annual reports from 2007 and 2008, <http://dataweb.usitc.gov> (accessed on January 26, 2010).

ⁱ Many market sources such as LME, Comex, Platts Metals Weekly, and others.

^j Private communication with USGS.

kWe. For thermal to electricity conversion, a Rankine cycle is the most practical and can produce electricity at greater than 25%

Table 2

Materials and fabrication costs of components constituting the boiler of the reaction system.

Component	Material	Unit cost (US \$) ^a	Fabrication/installation cost (US \$) ^b (%)	Total component cost (US \$)
Cell tubes	304 Stainless steel tubing 140 tubes of 29.2 cm ID, 1.76 m length, 6.35 mm wall thickness	55,785	100	111,570
Aluminum shot	5 mm diameter 14,283 lb	1.1 per lb ^c	100	31,709
Heat transfer tube	304 Stainless steel tubing 840 tubes of 1.91 cm ID, 2 m length, 0.32 cm wall thickness	8400	100	16,800
Insulation	20 m ³	7225	100	14,450
Total cost (US \$)				174,529

^a Tubing cost using <http://www.meps.co.uk/composite%20stainless%20steel%20priceindex.html> and verbal quote for A182 alloy steel from <http://www.aasteel.com>.^b Estimated at 100% cost of tube as per Chapter 9, Perry's Chemical Engineers' Handbook, 8th edition (2007), McGraw-Hill Professional, New York.^c Many market sources such as LME, Comex, Platts Metals Weekly, and others. December 2010.**Table 3**

Summary of costs of power system.

System component	System component cost (US \$ per kWt/kWe) ^a
Reactant mixture	17/68
Boiler	174/696
Balance of plant (Rankine cycle)	175/300 ^b
Total cost (US \$)	366/1064

^a 25% Thermal to electric conversion efficiency.^b Private communication with Conectiv and confirmed with "Cost and Performance Baseline for Fossil Energy Plants" DOE/NETL-2007/1281 Rev. 1 2008 PC Coal and NGCC Balance of Plant Estimates; All for 200–500 MW scale Rankine plants. Altran Solutions provided an estimate for a one of a kind, 350 kWe gross, ~250 kWe net that would add about \$1700 per kWe to the cost if considering a single DG unit.

efficiency with capital costs for the balance of a plant of about \$300 per kWe. Adding the cost of the chemicals and boiler to the cost of the conversion system yields a total cost of about \$1064 per kWe. A summary of the cost breakdown is given in Table 3.

Appendix A. Operational parameters of a continuous thermal power system

A.1. Reactor parameters

For 1 MW of thermal power output, the total chemical mass is $1000 \text{ (kW)}/0.67 \text{ (kW/kg)} = 1500 \text{ kg}$. The volume of chemical reactant is $1,500 \text{ kg}/(0.38 \text{ kg/l)} = 3947 \text{ l}$. Assuming 25% of the reactor volume is occupied by solid fuel, the total reactor volume is $3947 \text{ l}/0.25 = 15,800 \text{ l} = 15.8 \text{ m}^3$.

The internal cell pressure is determined by the reactant NaH of the mixture since it is the only chemical that can decompose and build up pressure in the cell at the reaction temperature of 550 °C. Each cell is designed for the equilibrium pressure of NaH at 550 °C of about 2.27 MPa. A cylindrical cell having a 30.5 cm OD, a 0.635 cm cylindrical wall thickness, and 3.81 cm thick end plates on both ends provides for safe operation of stainless steel according to pressure vessel design standards. To achieve 1 MW thermal, the cell dimensions and power output per cell vary with the number of cell as shown in Table 4. If we choose 140 cells, each cell is 176 cm long, and the power output of a single cell is 7.14 kW. The corresponding reactor mass is about 120 kg.

Table 4

Dimensions and power output per cell as a function of cell number to achieve 1 MW thermal.

Number of cells	Chemical volume (l)	Single cell volume (l)	Cell length (m)	Power/cell (kW)
20	197.37	789.47	11.86	50.00
40	98.68	394.74	5.97	25.00
60	65.79	263.16	4.00	16.67
80	49.34	197.37	3.02	12.50
100	39.47	157.89	2.43	10.00
120	32.89	131.58	2.04	8.33
140	28.20	112.78	1.76	7.14
160	24.67	98.68	1.55	6.25
180	21.93	87.72	1.39	5.56
200	19.74	78.95	1.25	5.00
220	17.94	71.77	1.15	4.55
240	16.45	65.79	1.06	4.17
260	15.18	60.73	0.98	3.85
280	14.10	56.39	0.92	3.57
300	13.16	52.63	0.86	3.33

A.2. Steam generation parameters

Given the fixed heat source, the optional steam generation parameters are selected to achieve efficiency and economy of design. Then, the heat transfer medium and heat collector from the cell to the steam side of the system is matched up in the heat transfer for steam generation section. The thermal power generated from the reactants is used to generate saturated steam at 360 °C. Fig. 4 shows the flow diagram of steam generation. Water at room temperature (about 25 °C) flows into a heat exchanger where it is mixed with saturated steam and heated to a saturated temperature of 360 °C by the condensation of steam. A booster pump increases the water pressure to a saturation pressure of 18.66 MPa at 360 °C. The saturated water flows through the heat transfer tubes of the water wall to generate steam at the same temperature and pressure. Part of the steam flows back to a heat exchanger to preheat incoming return water from a turbine, while part of it goes to the turbine to generate electrical power. Additionally, the non-evaporated water in the water wall is recirculated to maintain an even temperature along each heat transfer tube. To achieve this, a steam collection line receives heated coolant comprising steam and non-evaporated water and delivers it to a steam-water separator (Fig. 2). Water is pumped from the bottom section of the separator to return to the heat transfer tubes through a water distribution line. The steam flows from the top of the separator to the turbine with a fraction diverted to the heat exchanger to preheat the return water from the turbine.

The mass balance and energy balance equations based on Fig. 4 are

$$m_2 = m_1 + m_4 \quad (1)$$

$$m_4(h_{3,v} - h_{2,l}) = m_1(h_{2,l} - h_{1,l}) \quad (2)$$

$$Q = m_2(h_{3,v} - h_{2,l}) = m_1(h_{2,l} - h_{1,l}) \quad (3)$$

where m is mass flow rate in kg/s, h is enthalpy in kJ/kg, and Q is the total energy released from the reactants in kJ/s (kW). Subscripts v and l represent the vapor phase and liquid phase of water. From the steam tables [8], $h_{1,l} = 104.9$ kJ/kg, $h_{2,l} = 1760.5$ kJ/kg, and $h_{3,v} = 2481$ kJ/kg, respectively. For $Q = 1000$ kW thermal power, the parameters given by Eq. (15) below are $m_1 = 0.42$ kg/s, $m_2 = 1.39$ kg/s, and $m_4 = 0.97$ kg/s, respectively.

A.3. Heat transfer for steam generation

Fig. 5 shows the configuration of the cross-section of a single cell and water heat transfer tube for steam generation. The bottom

half of the cell is covered with insulation to force the energy transfer to the water wall. In order to maintain uniform heat flux to each heat transfer tube and provide efficient heat transfer from the cell top to the water wall, the top-half of cell is covered by a high thermally conductive material (TCM) such as copper or aluminum, and the heat transfer tubes are evenly distributed on the top of each cell and embedded in the high TCM. As shown in Fig. 5, the 30.5 cm OD cross-sectional span of each cell is covered by six, 2.54 cm OD heat transfer tubes with a thickness of 0.32 cm that are evenly spaced at 5.08 cm centers.

The heat transfer processes from the reactants to the steam include: (a) heat transfer from reactants to internal surface of the top-half of the cell involving the three parallel steps of H_2 natural convection, thermal radiation, and thermal conduction along the cell wall; (b) thermal conduction from the internal surface to the external surface of the top-half of the cell; (c) thermal conduction from the external surface of the top-half of the cell to the external wall of each heat transfer tube through the TCM; (d) thermal conduction from the external wall to the internal wall of each heat transfer tube; and (e) saturated boiling

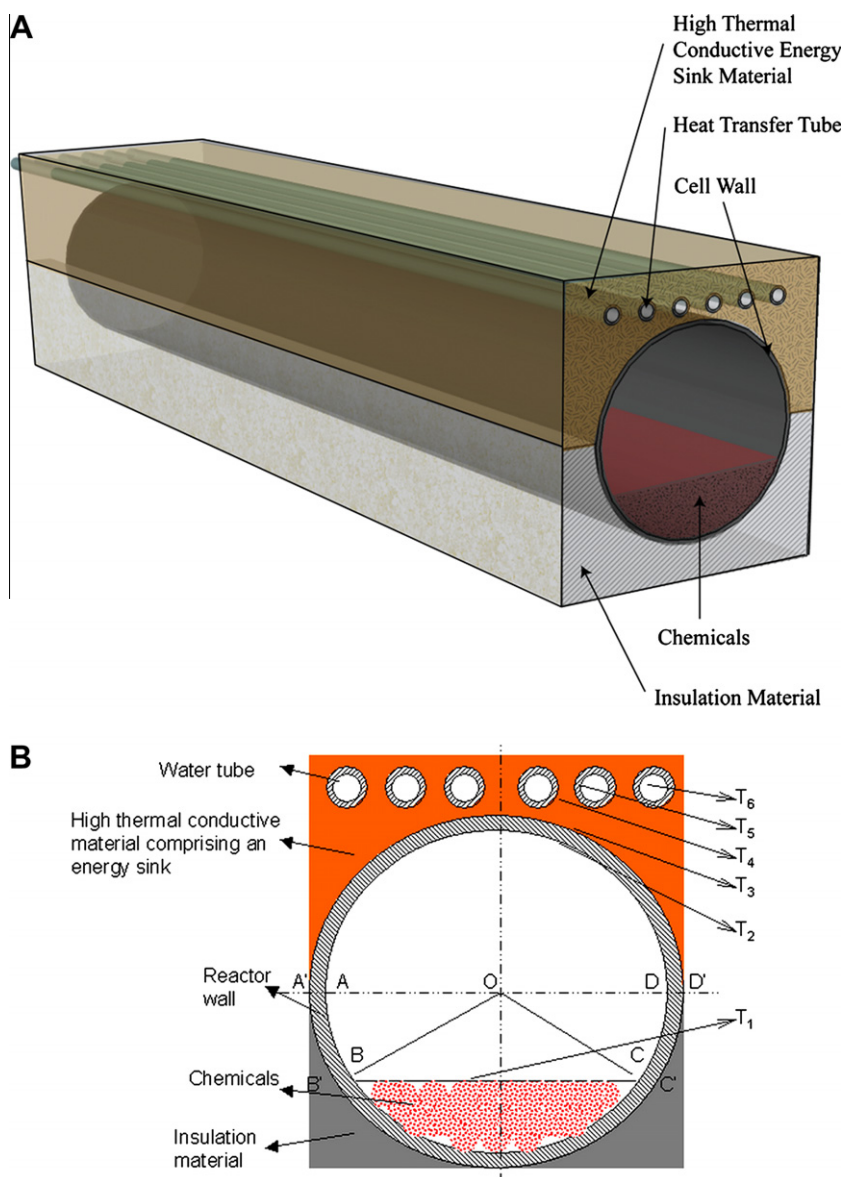


Fig. 5. Schematic of a reactor cell surrounded by insulation on the bottom and thermally conductive medium with embedded heat transfer tubes to accept the heat flow on the top. (A) Longitudinal view. (B) Cross-sectional view.

two-phase-flow heat transfer from the internal wall of each heat transfer tube to water.

A.3.1. Heat transfer from the reactants to the internal surface of the top-half of the cell

Consider heat transfer from the reactants to the internal surface of the top-half of the cell through H_2 natural convection, thermal radiation, and thermal conductivity along the cell wall. The dimensions of internal structure are: $\angle BOC = 2.31$ radians, width of BC $L_{BC} = 26.67$ cm, arc length $Arc_{AB} = Arc_{CD} = 6.07$ cm, $Arc_{A'B'} = Arc_{C'D'} = 6.35$ cm, $Arc_{BC} = 33.73$ cm, $Arc_{B'C} = 35.2$ cm, cell internal length $L = 168.3$ cm, and cell internal diameter $D = 29.21$ cm.

(i) H_2 natural convection. Heat transfer due to H_2 natural convection occurs from the solid fuel surface BC to the top-half circle AD. Assume that the reactant temperature T_1 is uniform and that the temperature T_2 of the half circle of the top internal surface of the reactor is uniform. Then, the empirical model for the free convection from a horizontal plate isothermally heated from the bottom in an enclosure gives the result that the Nusselt number is

$$Nu_{\delta_{H_2}} = \frac{ke_{H_2}}{k_{H_2}} = 0.061(Gr_{\delta_{H_2}} Pr_{H_2})^{\frac{1}{3}} \quad \text{if } Gr_{\delta_{H_2}} Pr_{H_2} > 3.2 \times 10^5 \quad (4)$$

where ke_{H_2} and k_{H_2} are the apparent thermal conductivity and molecular thermal conductivity of H_2 at 550 °C and 2.27 MPa, respectively, δ_{H_2} is the distance from hot surface to cold surface, Pr_{H_2} is the Prandtl number of H_2 , Gr is the Grashof number

$$Gr_{\delta_{H_2}} = \frac{\rho g \beta (T_1 - T_2) \delta_{H_2}^3}{\mu^2} \quad (5)$$

where ρ and μ are the density (~ 0.768 kg/m³) and viscosity ($\sim 1.59 \times 10^{-5}$ N s/m²) of H_2 , respectively, g is acceleration of gravity (9.8 m/s²) and $\beta = 2/(T_1 + T_2)$. Power flow by H_2 natural convection is

$$q_{1,H_2} = ke_{H_2} L_{BC} L \frac{T_1 - T_2}{\delta_{H_2}} \quad (6)$$

where the pseudo natural convection distance δ_{H_2} of 0.157 m can be calculated from

$$\delta_{H_2} = \frac{\pi D - L_{BC} - Arc_{BC}}{2} \quad (7)$$

(ii) Thermal radiation. By assuming a shape factor of 1 for the thermal radiation from the surface of the reactants to the half circle of the cell top internal surface, the power flow due to thermal radiation can be calculated from

$$q_{1,r} = \frac{\sigma(T_1^4 - T_2^4)}{\frac{1}{L_{BC}L} + \frac{1-\varepsilon}{\varepsilon\pi DL/2}} \quad (8)$$

where σ is the Stefan–Boltzmann constant ($= 5.669 \times 10^{-8}$ W/m² K⁴), and ε is emissivity of reactor material ($= 0.8$ for stainless steel). The reactants comprise a black material, which has emissivity equal to 1.

(iii) Thermal conductivity along the cell wall Arc BA and CD with the cell stationary. If the cell remains stationary and a scraper rotating inside the cell mixes the reactants, the thermal conductivity can be calculated using the equation:

$$q_{1,cs} = 2k_r \delta_r L \frac{T_1 - T_2}{(Arc_{AB} + Arc_{A'B'})/2} \quad (9)$$

where k_r is thermal conductivity of the cell (19 W/mK) and δ_r is cell thickness. Eq. (9) applies when the following assumptions are satisfied:

- Temperature across the cross-section of AA' and DD' is the same as T_2 . This assumption is reasonable since the conductivity of the high thermally conductive medium (TCM) such as copper is high, 363 W/mK compared to 19 W/mK for stainless steel at 400 °C, and the mass of the TCM is large.
- No heat transfer occurs vertical to the arc surface AB, A'B', CD, and C'D'. Since the bottom half of the cell is covered with insulation, no heat transfer occurs vertical to the arc surface A'B'C'D'. Although there is some energy transfer from the surface of the reactants to the internal surface along arc AB and CD, the energy absorbed is also radiated out to the top surface. Therefore, it is reasonable to assume that no net heat transfer occurs vertical to the arc surface AB and CD.

A.3.2. Thermal conduction from the top of the cell internal wall surface to the external surface

The thermal conduction from the top of the cell internal wall surface to the external surface can be calculated using the following equation

$$q_2 = \frac{\pi k_r L (T_2 - T_3)}{\ln(D_{RO}/D_{RI})} \quad (10)$$

where D_{RO} and D_{RI} are external diameter and internal diameter of the cell, respectively.

A.3.3. Thermal conduction from the thermally conductive medium to the external wall of heat transfer tube

The thermal conduction from the high TCM to the external wall of heat transfer tube can be calculated by the following equation

$$q_3 = \frac{k_{ES} D_{RO} L (T_3 - T_4)}{\delta_{ES}} \quad (11)$$

where k_{ES} and δ_{ES} are the thermal conductivity and the pseudo thickness of the TCM, respectively. The latter can be calculated as

$$\delta_{ES} = \frac{D_{RO} \cdot (D_{RO}/2 + D_{WO}) - \pi D_{RO}^2/8 - n_W \pi D_{WO}^2/4}{D_{RO}} \quad (12)$$

where D_{WO} is external diameter of the heat transfer tube and n_W is the number of heat transfer tubes for each cell.

A.3.4. Thermal conduction from the external wall to the internal wall of the heat transfer tube

The thermal conduction from the external wall to the internal wall of the heat transfer tube can be calculated by

$$q_4 = \frac{2\pi n_W k_W L (T_4 - T_5)}{\ln(D_{WO}/D_{WI})} \quad (13)$$

where k_W and D_{WI} are the thermal conductivity (19 W/mK) and the internal diameter of each heat transfer tube, respectively.

A.3.5. Saturated boiling two-phase flow heat transfer from the internal wall of the heat transfer tube to the water

The saturated boiling two-phase flow heat transfer from the internal wall of the heat transfer tube to the water can be calculated as

$$q_5 = n_W h_{TP} \pi D_{WI} L (T_5 - T_6) \quad (14)$$

where the two-phase heat transfer coefficient h_{TP} can be calculated by relationship after Kandlikar [9] as

$$\frac{h_{TP}}{h_l} = C_1 Co^{C_2} (25 Fr_{lo})^{C_3} + C_3 Bo^{C_4} F_{fl} \quad (15)$$

where h_l is the single-phase heat transfer coefficient with only the liquid fraction flowing in the tube:

$$h_l = 0.023 \text{Re}_l^{0.8} \text{Pr}_l^{0.4} \left(\frac{k_l}{D_{wl}} \right) \quad (16)$$

where Re_l and Pr_l are the water Reynolds number and Prandtl number, respectively:

$$\text{Re}_l = \frac{GD_{wl}(1-x)}{\mu_l} \quad (17)$$

$$\text{Pr} = \frac{C_p \mu}{k} \quad (18)$$

Bo in Eq. (15) is the boiling number, which can be calculated as

$$Bo = \frac{q}{G i_{lg}} \quad (19)$$

and Co is the convection number that can be calculated as

$$Co = \left(\frac{1-x}{x} \right)^{0.8} \left(\frac{\rho_g}{\rho_l} \right)^{0.5} \quad (20)$$

Fr_{lo} is the Froude number with all flow as liquid given by

$$\text{Fr}_{lo} = \frac{G^2}{\rho_l^2 g D} \quad (21)$$

In Eqs. (15)–(21), x is the dryness fraction, G is mass flux in $\text{kg/m}^2 \text{s}$, μ is dynamic viscosity in N s/m^2 , ρ is density in kg/m^3 , k is thermal conductivity in W/mC , i_{lg} is latent heat of vaporization in J/kg , C_p is specific heat in J/kg C , g is acceleration of gravity in m/s^2 , and q is heat flux in W/m^2 . Subscripts l and g represent the liquid and vapor phase, respectively. C_1 – C_5 are constants given in Table 5, in which $Co < 0.65$ corresponds to the convective boiling region and $Co > 0.65$ corresponds to the nucleate boiling region.

The calculated two-phase heat transfer coefficient h_{TP} as a function of the recirculation ratio, R , predominantly determines the water flow rate in the heat transfer tube.

Table 5
Constants used in Eq. (15).

Constant	Convection region	Nucleate boiling region
C_1	1.136	0.6683
C_2	−0.9	−0.2
C_3	667.2	1058.0
C_4	0.7	0.7
C_5	0.3	0.3

Consider the flow scheme shown in Fig. 4 and the partition of the coolant between liquid and gaseous phase. If the saturated water mass flow rate in the boiler is 1.39 kg/s, the water would be completely converted to steam by the end of the heat transfer tube which would lead to a low heat transfer rate. The absence of water in the heat transfer tube end can be avoided by maintaining saturated water recirculating in each heat transfer tube. In the case that the cycle ratio is 1 (1/2 of the coolant is recycled as water and 1/2 is output as steam), the actual saturated water flow rate must be 2.78 kg/s in the heat transfer tube, and the total steam flow rate from the 140-cell system is 1.39 kg/s. Then, the heat transfer coefficient calculated from Eq. (15) is $2.17 \text{ kW/m}^2 \text{ } ^\circ\text{C}$.

A suitable design that provides a high degree of control over the operating temperatures and maintains the regeneration and power reaction comprises stationary (non-rotating) reactors with the reactants mixed by rotation of an internal scraper. The desired temperature profile in the boiler system having an inter-reactor gradient and an external gradient between the top cell wall and the heat transfer tubes embedded in the TCM may be readily achieved by adjusting the thickness and material selection based on thermal conductivity of the insulation at the top of the cell, the TCM, and the heat transfer tube insulation. Assuming the rotation of the internal scraper does not enhance heat transfer, the calculated temperature profiles are shown in Table 6. Using copper or aluminum shot as the TCM, a reaction temperature of $550 \text{ } ^\circ\text{C}$ is difficult to maintain with 1 MW thermal power output. By using stainless steel as the TCM, the reactant temperature would rise to $532 \text{ } ^\circ\text{C}$, but a large temperature gradient would exist within the TCM ($402 \text{ } ^\circ\text{C}$ vs. $367 \text{ } ^\circ\text{C}$). The gradient would cause an uneven heat transfer rate in the heat transfer tubes. This gradient can be corrected using the copper TCM with the addition of a 0.597 cm thick layer of insulation material such as ceramic fiber on the external surface of each heat transfer tube or the addition of a 0.498 cm thick layer of ceramic fiber on the top external surface of each cell. Either modification maintains a reactant temperature of $550 \text{ } ^\circ\text{C}$. Alternatively, aluminum is a cheaper replacement for copper. Selecting 5 mm aluminum shot as the TCM with the addition of a 0.564 cm thick layer of ceramic fiber on the external surface of each heat transfer tube or the addition of a 0.472 cm thick layer of ceramic fiber on the top external surface of each cell would maintain a reactant temperature of $550 \text{ } ^\circ\text{C}$. Therefore, it is feasible to continuously release thermal power of 1 MW at a reactant temperature of $550 \text{ } ^\circ\text{C}$.

Table 6
The effect of adding insulation to the cell top and heat transfer tubes on the internal and external temperature profiles. The temperatures refer to those designated in Fig. 5.

Thermally conductive material (TCM)	Aluminum	Aluminum + ceramic fiber	Aluminum + ceramic fiber	Copper	Copper + ceramic fiber	Copper + ceramic fiber	Stainless steel
Solid fuel temperature, T_1 ($^\circ\text{C}$)	508.60	550.00	550.00	506.63	550.00	550.00	532.15
Cell internal top wall temperature, T_2 ($^\circ\text{C}$)	374.65	428.79	428.63	372.01	428.78	428.85	405.46
Cell external top wall temperature, T ($^\circ\text{C}$)	371.63	425.77	425.61	368.99	425.76	425.83	402.43
Insulation thickness (cm)			0.472			0.498	
Insulation internal temperature ($^\circ\text{C}$)			425.61			425.83	
TCM bottom temperature, T_3 ($^\circ\text{C}$)	371.63	425.77	371.63	368.99	425.76	368.99	402.43
TCM top temperature, T_4 ($^\circ\text{C}$)	367.16	421.30	367.16	367.16	423.93	367.16	367.16
Insulation thickness (cm)		0.564			0.597		
Insulation external temperature ($^\circ\text{C}$)		421.30			423.93		
Water wall external temperature, T_4 ($^\circ\text{C}$)	367.16	367.16	367.16	367.16	367.16	367.16	367.16
Water wall internal temperature, T_5 ($^\circ\text{C}$)	365.45	365.45	365.45	365.45	365.45	365.45	365.45
Water temperature, T_6 ($^\circ\text{C}$)	360	360	360	360	360	360	360

Appendix B. Effect of cell undergoing rotation on the internal and external temperature gradients

If the cell undergoes rotation at n rotations per minute (rpm), the heat transfer from reactants to the TCM at the cell top involves the standard thermal conductivity processes as well as the exchange of heat between the reactants and the cell wall with latter heat transferred by physical transport. If the cell rotates counterclockwise, the cell wall ABB'A' rotates down to contact the reactants and absorbs energy causing its temperature to rise. Whereas, with a clockwise rotation, the opposite occurs, the reactants absorb heat from the cell wall, and their temperature drops due to thermal conduction. To model this component, the steady-state temperature profile in the cell wall ABCDD'C'B'A' is determined by solving the following partial differential equation:

$$\rho_r C_p \frac{n}{60r} \frac{\partial T}{\partial \theta} = k_r \left(\frac{\partial^2 T}{\partial r^2} + \frac{1}{r} \frac{\partial T}{\partial r} + \frac{1}{r^2} \frac{\partial^2 T}{\partial \theta^2} \right) \quad (22)$$

wherein the boundary conditions are

- for the external surface along arc A'B'C'D': $\frac{\partial T}{\partial r} = 0$ ($\theta = 0 \sim \pi$, $r = 15.24$ cm)
- for the internal surface along arc AB and CD: $\frac{\partial T}{\partial r} = 0$ ($\theta = 0 - 0.415$ radians and 2.725 radians $\sim \pi$, $r = 14.61$ cm)
- for the internal surface along arc BC: $T = T_1$ ($\theta = 0.415 - 2.725$, $r = 14.61$ cm)
- for the cross-section of AA' and DD': $T = T_2$ ($\theta = 0$ and π , $r = 14.61 - 15.24$ cm)

where ρ_r and C_p are the density and heat capacity of the reactor material, respectively. The solution of the partial differential Eq. (22) can be simplified by separating arc ABCDD'C'B'A' into the three arc sections ABB'A', BCC'B', and CDD'C'. The corresponding simplified differential equations are then solved separately.

The analysis of the heat transfer processes from the solid fuel to arc section BCC'B' of the cell can be simplified by modeling the heat transfer as being equivalent to that of a flat plate with thickness of δ_r , as shown in Fig. 6. The boundary conditions for arc section BCC'B' are

- the thermal conduction occurs only in the radial direction;
- once the cell wall contacts the chemicals, the internal wall temperature arrives at temperature T_1 and stays there until the position of the internal wall advances from contact with the chemicals;
- the rotation rate is constant.

The heat transfer equation to describe the temperature distribution in the cell wall is

$$\frac{\partial T(z, t)}{\partial t} = \frac{k_r}{\rho_r \cdot C_p r} \frac{\partial^2 T(z, t)}{\partial z^2} = \alpha_r \frac{\partial^2 T(z, t)}{\partial z^2} \quad (23)$$

with the following boundary conditions

$$T(0, t) = T_1 \quad (24)$$

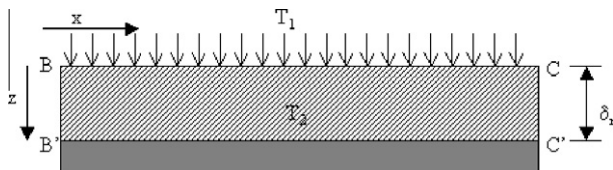


Fig. 6. The simplification of the heat transfer in the arc BCC'B' as being equivalent to the heat transfer in a flat plate.

$$\frac{\partial T(\delta_r, t)}{\partial z} = 0 \quad (25)$$

and the initial condition

$$T(z, 0) = T_2 \quad (26)$$

Solving Eq. (23) gives the temperature distribution in the cell wall:

$$T(z, t) = T_1 - (T_1 - T_2) \sum_{p=0}^{\infty} \frac{2}{(p+0.5)\pi} \times \sin \left[\frac{(p+0.5)\pi z}{\delta_r} \right] \exp \left[-\frac{(p+0.5)^2 \pi^2 \alpha_r t}{\delta_r^2} \right] \quad (27)$$

Since the relationship between time t and distance x with a rotation rate of $n2\pi$ radians per minute is

$$t = \frac{x}{n\pi D} \quad (28)$$

the temperature distribution along arc BCC'B' is

$$T(z, x) = T_1 - (T_1 - T_2) \sum_{p=0}^{\infty} \frac{2}{(p+0.5)\pi} \times \sin \left[\frac{(p+0.5)\pi z}{\delta_r} \right] \exp \left[-\frac{(p+0.5)^2 \pi \alpha_r x}{n\delta_r^2 D} \right] \quad (29)$$

The heat transfer rate along arc BC is

$$\begin{aligned} dq_x &= -k \cdot dA \cdot \frac{\partial T}{\partial z} \Big|_{z=0} = -k_r \cdot L \cdot dx \cdot \frac{\partial T}{\partial z} \Big|_{z=0} \\ &= \frac{2k_r L (T_1 - T_2)}{\delta_r} \sum_{p=0}^{\infty} \exp \left[-\frac{(p+0.5)^2 \pi \alpha_r x}{nD\delta_r^2} \right] dx \end{aligned} \quad (30)$$

The integration of Eq. (30) gives the heat flow from the chemicals to the cell wall

$$\begin{aligned} Q_x &= \int_0^{Arc_{BC}} dq_x \\ &= \frac{2k_r L (T_1 - T_2)}{\delta_r} \sum_{p=0}^{\infty} \int_0^{Arc_{BC}} \exp \left[-\frac{(p+0.5)^2 \pi \alpha_r x}{nD\delta_r^2} \right] dx \\ &= \frac{2k_r L (T_1 - T_2) n\delta_r D}{(p+0.5)^2 \pi \alpha_r} \left\{ 1 - \sum_{p=0}^{\infty} \exp \left[-\frac{(p+0.5)^2 \pi \alpha_r}{n\delta_r^2 D} \cdot Arc_{BC} \right] \right\} \end{aligned} \quad (31)$$

The thermal-conduction heat flow from the chemicals to the cell wall as a function of the cell rotation rate (Eq. (31)) with the boundary conditions of $T_1 = 550^\circ\text{C}$ and $T_2 = 400^\circ\text{C}$ is shown in Fig. 7.

For arc section CDD'C', the boundary conditions are

- the thermal conduction occurs only in the arc direction;
- the temperature at the cross-section of CC' is T_1 , the average temperature of T_1 and the temperature at C', and the temperature at the cross-section of DD' is T_2 ;
- no heat transfer occurs on the surfaces of the internal and external walls.

The heat transfer from the reactor cross-section CC' to D'D' can be simplified as being equivalent to the heat transfer in the flat plate with thickness of Arc_{CD} shown in Fig. 8. The heat transfer equation to describe the temperature distribution in the reactor wall is given by

$$\rho_r \cdot C_p \cdot u \frac{\partial T(x)}{\partial x} = k_r \frac{\partial^2 T(x)}{\partial x^2} \quad (32)$$

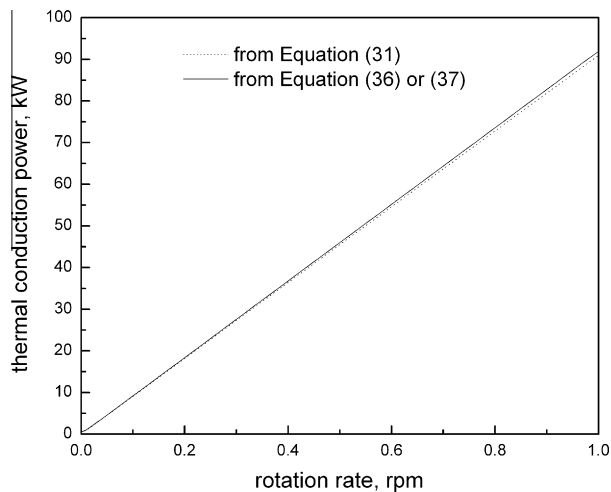


Fig. 7. The thermal-conduction heat flow from the chemicals to the cell wall as a function of the cell rotation rate (Eq. (31)) with the boundary conditions of $T_1 = 550^\circ\text{C}$ and $T_2 = 400^\circ\text{C}$.

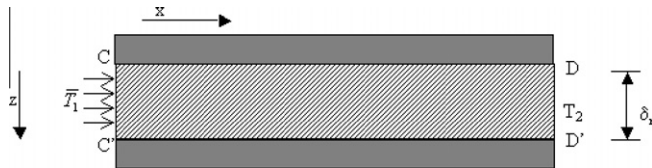


Fig. 8. The simplification of the heat transfer in the arc CDD'C' as being equivalent to the heat transfer in the flat plate.

where u is the linear velocity πnD when the cell is rotating. Setting

$$B = \frac{\pi D \rho_r C_{p_r}}{k_r} \quad (33)$$

Eq. (32) becomes

$$\frac{\partial^2 T(x)}{\partial x^2} = nB \frac{\partial T(x)}{\partial x} \quad (34)$$

Solving Eq. (34) gives

$$T = \bar{T}_1 + \frac{\exp(nBx) - 1}{\exp(nB \cdot \text{Arc}_{CD}) - 1} (T_2 - \bar{T}_1) \quad (35)$$

The heat transfer from cross-section CC' to DD' can be calculated by

$$\begin{aligned} q|_{x=0} &= \pi n D \rho_r C_{p_r} \delta_r L (\bar{T}_1 - T_2) - k_r \delta_r L \left. \frac{dT}{dx} \right|_{x=0} \\ &= \pi n D \rho_r C_{p_r} \delta_r L (\bar{T}_1 - T_2) + \frac{n B k_r \delta_r L (\bar{T}_1 - T_2)}{\exp(nB \cdot \text{Arc}_{CD}) - 1} \end{aligned} \quad (36)$$

which can also be calculated by

$$\begin{aligned} q|_{x=L} &= \pi n D \rho_r C_{p_r} \delta_r L (T_2 - \bar{T}_1) - k_r \delta_r L \left. \frac{dT}{dx} \right|_{x=L} \\ &= \frac{n B k_r \delta_r L (\bar{T}_1 - T_2)}{\exp(nB \cdot \text{Arc}_{CD}) - 1} \cdot \exp(nB \cdot \text{Arc}_{CD}) \end{aligned} \quad (37)$$

The thermal-conduction heat flow from the chemicals to the cell wall as a function of the cell rotation rate (Eqs. (36), (37)) with the boundary conditions of $T_1 = 550^\circ\text{C}$ and $T_2 = 400^\circ\text{C}$ is also shown in Fig. 7. The results from Eqs. (36), (37) agree with those from Eq. (31) and indicate that the thermal model gives reasonable predictions for the series heat transfers from the chemicals to the cell wall arc BC and between the cell wall cross-sections CC' and DD'. Any energy transferred from the chemicals to the cell wall arc BC is conserved by flowing to the TCM comprising an energy sink at the cell top through wall CDD'C' as shown in Figs. 6 and 8.

Fig. 7 shows that the heat flow increases very substantially with increasing rotation rate. For example, a rotation rate of 0.08 rpm would cause the total power of 7.14 kW released from reactants in each cell (Table 4) to be transferred to the energy sink by mass transport and thermal conduction. An excessive rotation rate will cause the temperature of the chemicals to rapidly drop, and this would effect the performance of reactants to release energy. An acceptable rotation rate of 0.08 rpm may not provide sufficient mixing of the chemicals. Then, with the cell remaining stationary, a scraper inside the cell may be a better option to achieve mixing of the regenerated component with the other reactants of the reaction mixture.

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